DOE/EP-0004

# Assessment of Costs and Benefits of Flexible and Alternative Fuel Use in the U.S. Transportation Sector

## Technical Report Eleven: Evaluation of a Potential Wood-to-Ethanol Process

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2601

This report is based on a study that Chem Systems, Inc., prepared for the National Renewable Energy Laboratory, formerly the Solar Energy Research Institute.

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## *EXECUTIVE SUMMARY*

In 1988 the Department of Energy (DOE) initiated a comprehensive technical analysis of an alternative-fuel transportation system in the United States. During the next two decades, alternative fuels such as alcohol (methanol or ethanol), compressed natural gas (CNG), liquefied petroleum gas (LPG), and electricity could become practical alternatives to oil-based fuels in the U.S. transportation sector.

DOE is using the Alternative Fuels Trade Model (AFTM) to provide an integrating framework for its assessment of alternative fuels. The AFTM focuses on the production and consumption of alternative transportation fuels as substitutes for motor gasoline and diesel fuel. The AFTM determines prices and quantities that balance the interrelated world oil and gas markets. A critical modeling issue relates to the extent of market power held by the major oil-exporting nations and the manner in which such market power may be exercised. The AFTM model is sufficiently flexible to allow for the calculation of market balances under a variety of alternative characterizations of the world oil market. It characterizes market balances, or equilibria, in a selected year for multiple fuels that derive from oil or gas. The model is being used to examine the alternative fuels in several hypothetical multifuel scenarios. The supplies of the two principal raw materials (crude oil and natural gas) and the supplies of grain and cellulosic feedstocks for ethanol are represented by upward-sloping price-responsive curves. The model provides for fuel transportation between regions and includes processes that convert crude oil or natural gas to industrial and consumer fuels and that convert grain or cellulosic feedstocks to ethanol. The AFTM models the final demand for each fuel by downward-sloping constant-elasticity demand curves. It provides opportunities for long-run fuel substitution in flexible-fuel vehicles and industrial and utility boilers. The degree of fuel switching by flexible-fuel vehicles influences the market penetration and success of alternative transportation fuels, such as methanol or CNG. Substitution between oil and gas in the industrial-utility boiler-fuel market establishes an important connection between the prices of petroleum products and gas-based products.

The AFTM provides insights into the market effects of introducing alternative transportation fuels. It estimates changes in the prices, supplies, and demands of conventional fuels. It reports the levels of alternative-fuel use and tracks the geographic sources of U.S. energy supplies. The economic costs and benefits of introducing these substitute fuels are also measured, based on a standard social surplus analysis. Net benefit is estimated as the benefits that consumers gain from their levels of final demand, minus all the costs of fuel production, transportation, and conversion.

To keep interested parties informed about the progress of the DOE Alternative Fuels Assessment, the Department periodically publishes reports dealing with particular aspects of this complex study. This report provides a technical and economic evaluation of a new process under development for producing ethanol from wood.

Interest in using ethanol as an oxygenate, octane enhancer, fuel extender, or neat liquid fuel has grown over the years. Widespread use of ethanol as a transportation fuel could reduce oil imports, slow the depletion of U.S. petroleum resources, and produce environmental gains.

Until now, the barrier to widespread ethanol use has been the lack of appropriate technology that would reduce the cost of ethanol to a reasonable level. Over the last 5 years, the National Renewable Energy Laboratory (NREL) (formerly the Solar Energy Research Institute) has developed a new process that incorporates recent significant improvements in the ethanol lignocellulosic wood process. NREL has proposed use of lignocellulosic materials (such as wood) to produce ethanol because of their low cost and their availability. Use of a renewable feedstock source such as wood could be a

long-term solution to the problem of dwindling petroleum reserves, provided no fossil fuel inputs are used during biomass production, harvesting, and transport. It can also be argued that use of ethanol from lignocellulose will result in no net contribution of carbon dioxide ($CO_2$) to the atmosphere. This is because the $CO_2$ released during biomass conversion to ethanol and ethanol combustion will be absorbed during the growth of new biomass materials to replace those used during conversion.

As part of its ongoing program, NREL has developed a conceptual process design for a large-scale production plant, based primarily on experimental data, to determine the economic feasibility of such a production plant. A detailed technical and economic analysis of this design has led to the following conclusions:

- The overall process concept appears to be feasible and is generally supported by NREL and related laboratory data as reasonable engineering judgment.
- The next step in the process development and scaleup program needs to be the construction of a pilot-scale plant with all process steps integrated to verify data and engineering assumptions, especially for a commercial-scale plant.
- Vendor laboratory experiments are necessary to verify the feasibility of large-scale equipment (for example, the disc refiner, impregnator, and prehydrolysis reactor).
- Based on the current design, the economics for the production of the ethanol are much improved over previous (mid-1980's) designs. At the base-case wood feed rate (1,920 dry short tons per day), wood cost ($42 per dry ton), and overall yield[1] (68 percent), the price of ethanol is $1.27 per gallon (including capital charges of 20 percent).
- According to initial laboratory results, improved overall yields are feasible. Many possibilities for yield improvements have been proposed. Assuming that the necessary research and development efforts continue and that these yield improvements are proven, the cost of ethanol production from wood could be reduced significantly. For example, if overall yield can approach 90 percent, at this point the price of ethanol can be reduced to $0.965 per gallon at the 1,920 dry tons per day wood feed case. This assumes constant investment and wood cost, including capital charges of 20 percent.
- If an analysis is made for a large plant (5 times the base-case wood feed capacity) then the ethanol price is estimated to be $1.02 per gallon. If one assumes the same yield improvements as above, then the ethanol price is reduced to $0.781 per gallon. Both values include capital charges of 20 percent.

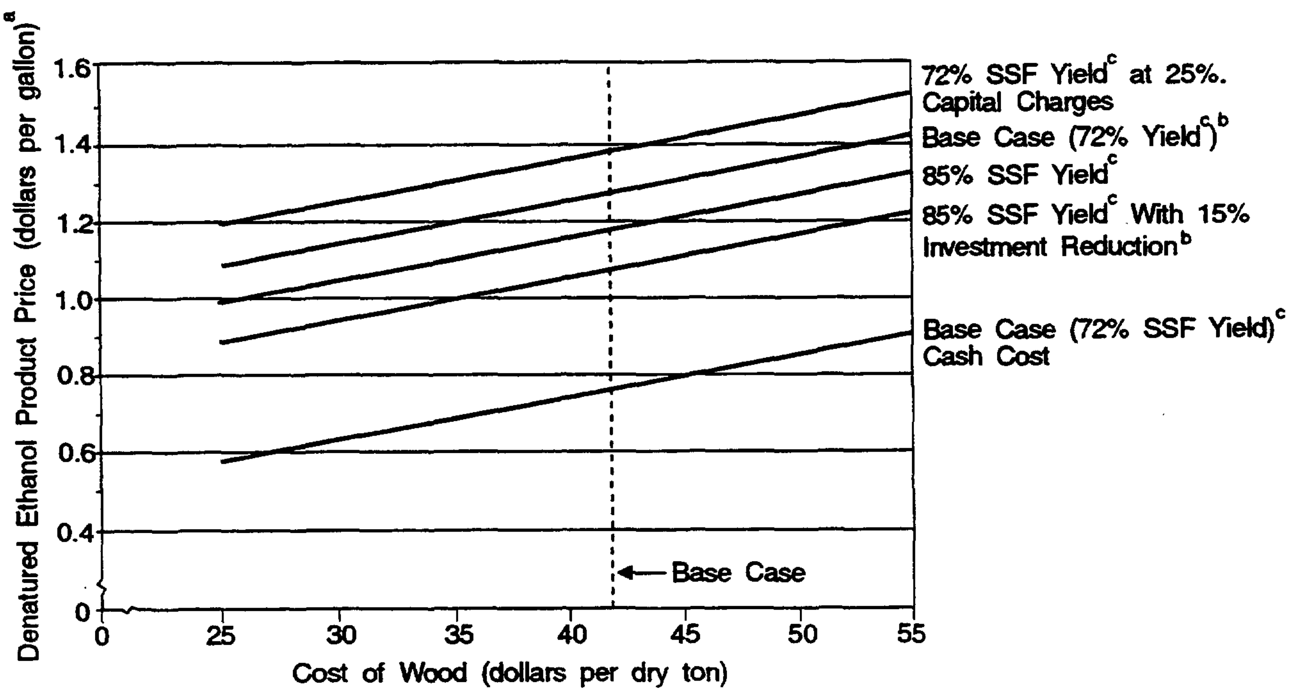
The effect of wood cost on ethanol price for various cases is illustrated in Figure S–1. Using the case of a large plant at 90 percent overall yield and a wood cost of $34 per dry ton, which according to NREL is the production goal established by Oak Ridge National Laboratory for energy crops, the ethanol price is reduced to $0.714 per gallon.

Figure S–2 shows the effect of improved yield on ethanol price for the base-case wood feed rate and the large plant, including a capital investment sensitivity. With wood at $42 per dry ton and overall yield at 90 percent, and assuming a large plant with 15 percent investment reduction, the ethanol price would be $0.732 per gallon.

If, in addition to the above improvements (yield, plant size, capital reduction, and feedstock cost), efforts are made to reduce power consumption, optimize other aspects of the process, and increase the carbohydrate content of the feedstock, then the ethanol price could possibly be reduced even further.

[1]Overall yield is defined as the mass fraction of cellulose and hemicellulose converted into ethanol.

**Figure S–1 — Effect of Wood Cost on Ethanol Product Price at Varying Conditions, 1,920 Dry Tons per Day Wood Feed**

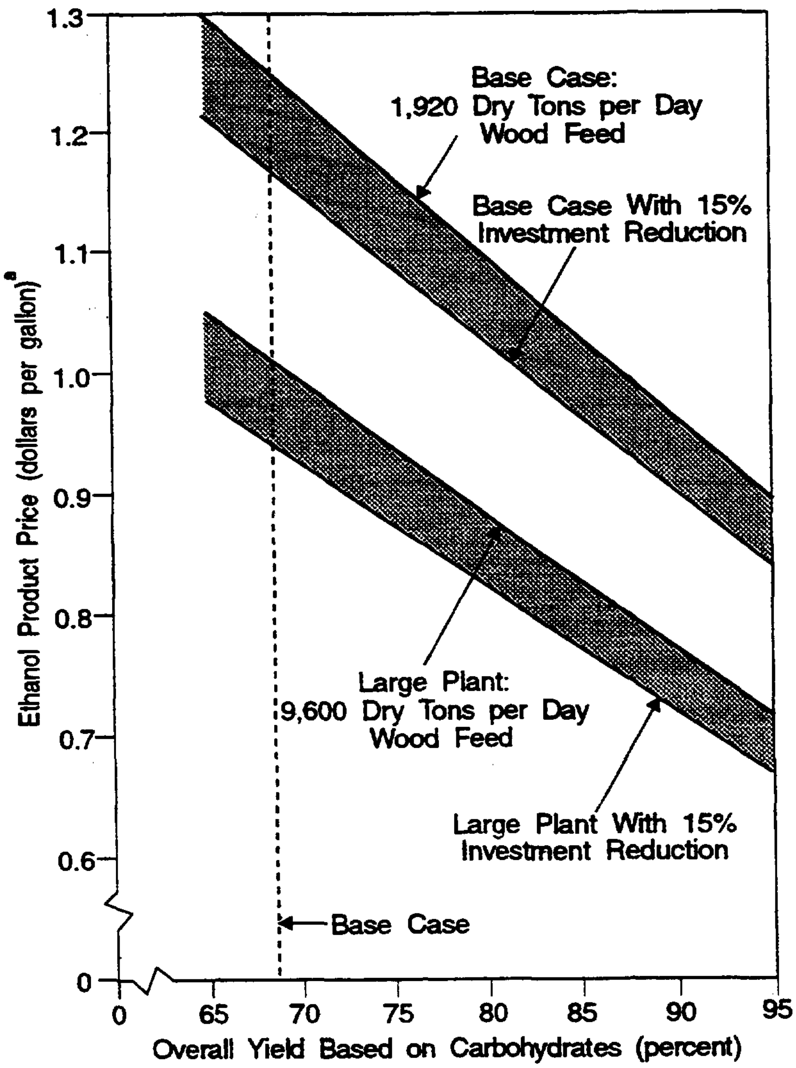



[a] Cost basis is U.S. Gulf Coast, last quarter of 1987 for denatured ethanol product.

[b] Includes annualized capital charges at 20% of capital cost.

[c] SSF yield is the mass fraction of cellulose entering the main simultaneous saccharification and fermentation (SSF) unit converted to ethanol. Overall yield is slightly lower than SSF yield because some cellulose is consumed prior to SSF as a carbon source for cellulase production and cell growth. In addition, hemicellulose conversion is lower than cellulose conversion.

**Figure S-2 — Effect of Overall Yield on Ethanol Product Price at Varying Conditions**




Note: Assumes $42 per dry ton of wood and 20 percent capital charges.

[a]U.S. Gulf Coast, last quarter of 1987.

# I. PROCESS EVALUATION

### *INTRODUCTION*

The National Renewable Energy Laboratory (NREL), formerly the Solar Energy Research Institute, has proposed a process for converting wood to ethanol. This process is a simplified, straightforward one that incorporates significant improvements over processes developed in the early 1980's. The major improvements lie in the following areas:

- Xylose fermentation to ethanol.
- Simultaneous saccharification and fermentation.
- Elimination of numerous separation and concentration steps.

In previous designs, the xylan component of wood was hydrolyzed to xylose and then converted to furfural. Although a byproduct credit was given to furfural, widespread ethanol production in the long term would result in a glut on the market, and the value of furfural as a byproduct would become questionable. Today, NREL has a process that can ferment xylose to ethanol while reducing the amount of xylose converted to furfural. Xylose fermentation by itself can increase the production of ethanol by 25 percent through the increased yield of ethanol.

The simultaneous saccharification and fermentation (SSF) process has several advantages over the earlier separate hydrolysis and fermentation (SHF) process. The key advantage is in the reduction in end-product inhibition of the cellulase enzyme complex at high glucose concentrations. This no longer occurs because the glucose that is formed in an SSF reactor is converted to ethanol very quickly and therefore does not build up in concentration.

This lack of inhibition allows for greatly reduced enzyme loading (from 33 to 7 IU per gram of cellulose, with an IU being an international unit of enzyme activity); this cuts the cost of enzyme production dramatically.

Another improvement lies in the unit operations of the process. As an example, several costly separation steps have been completely eliminated. Gypsum is no longer separated after neutralization, but only after ethanol distillation. Likewise, lignin flows from one process step to another and is removed only during the first stage of ethanol distillation. There is no multiple-effect evaporation of sugar solutions before fermentation, nor is there any furfural production step.

In general, the data on which the design conditions are based come from NREL laboratories. The reported yields are not the best ever achieved, but rather are conservative and reproducible values that form a very reasonable basis for design. The yields are not optimum values but rather a snapshot in time reflecting the current state of process development. Improvements are expected as research and development continues.

The major drawback in the design basis is the lack of actual experimental data from running an integrated process (that is, running all the process steps in series using effluent from one step as the feed to the next step). NREL plans to run an integrated process in the near future to demonstrate that the process will run as proposed.

### *STEPS IN THE CONVERSION PROCESS*

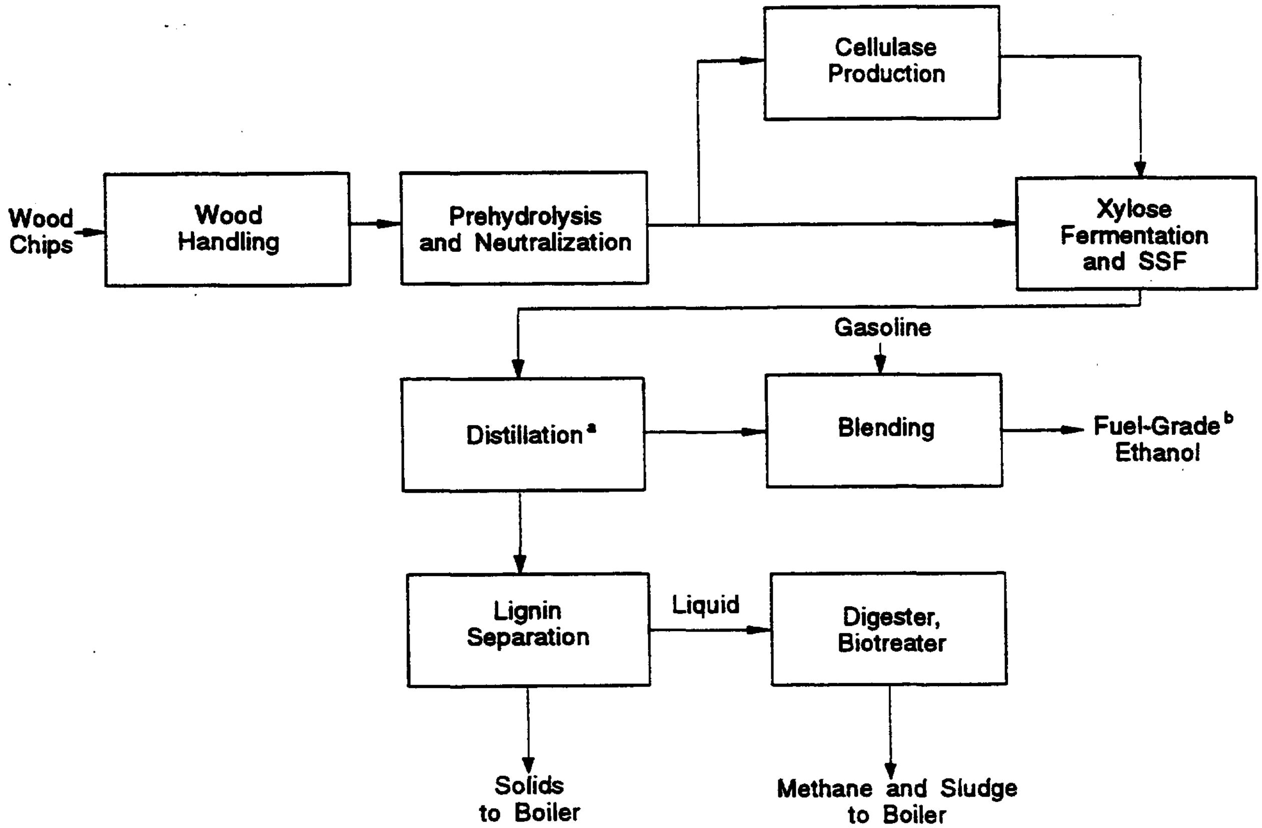
The following sections describe the NREL wood-to-ethanol conversion process. Figure I–1 outlines the main process units and flows. The description of each step in the process contains the design basis, a brief description, and comments on potential problems or possibilities.

#### Wood Handling and Size Reduction (Section 100)

##### Design Basis

The composition of the wood used in the material balance is a typical hardwood. Its composition is described in Table I–1.

## Figure I-1 — Overview of Wood-to-Ethanol Process




[a] Separation to azeotropic ethanol.

[b] 90.3 percent (of weight) ethanol, 4.7 percent water, 5 percent gasoline.

The feed rate is 160,000 pounds of dry wood per hour. Wood chips are assumed to contain 50 percent water.

Wood chips, approximately 1 inch in size, are delivered to the plant in 23-ton trucks. An outside contractor will deliver the wood chips on a schedule of one or two shifts per day, 5 days a week. Enough wood chips to allow 2 weeks of processing will be stored on site.

### Process Description

Figure I-2 depicts the wood handling and size reduction process. Wood chips are offloaded to a washing flume from three separate receiving stations with scales. The chips are transported from the water to a wood-chip pile via stacking conveyor and are then fed to a disc refiner, which reduces their size to 1 to 3 millimeters (0.04 to 0.12 inches).

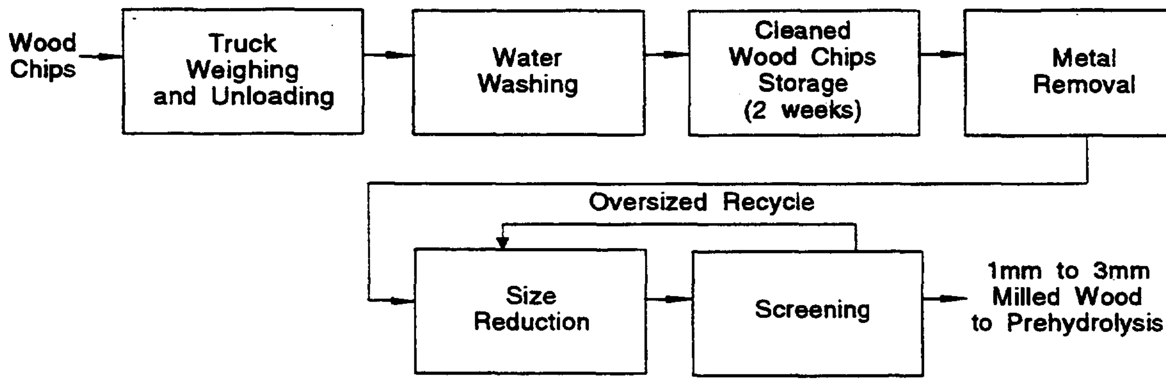
## Table I-1 — Composition of Wood Used in Process

| Component | Percentage of Weight |
|---|---|
| Cellulose | 46.2 |
| Xylan | 24.0 |
| Lignin | 24.0 |
| Solubles | 5.6 |
| Ash | 0.2 |
| Total | 100.0 |

### Comments

Procurement of wood chips is a complex operation and will require a dedicated system that includes logging, debarking, chipping,

**Figure I-2 — Section 100: Wood Handling and Size Reduction**




handling, offsite storage, and transportation to the ethanol plant on a large scale. Achieving a reliable wood-chip delivery system at a reasonable cost is crucial to efficient operation of the ethanol plant.

Approximately one-third of the power requirement for the entire plant is used in the chip-milling operation. The original NREL design incorporated three knife mills. However, analysis found that these mills do not have the capacity required for a reasonable design. Instead, four disc refiners (for example, Sprout-Bauer Model 45–1B) will be employed; horsepower requirements have been adjusted accordingly. It is important to run trials using this system to verify process parameters for the desired product size, including various methods of recycling oversized chips.

## Prehydrolysis and Neutralization (Section 200)

### Design Basis

The values for yields are based on NREL laboratory and pilot tests. Residence time and temperature have been adjusted to maximize xylan-to-xylose conversion; the system is not yet optimal. The design conditions for the prehydrolysis step are listed in Table I–2.

### Process Description

Figure I–3 depicts the prehydrolysis and neutralization process. Milled chips from the disc refiner are fed into a screw feeder that feeds the wood into an impregnator, where live steam and dilute sulfuric acid are injected. The residence time is 10 minutes at 100 degrees Celsius. The impregnator discharges the wood through a rotary valve to the prehydrolysis reactor. Live steam under pressure is injected into this reactor to heat the material up to reaction temperature under 6 atmospheres of pressure. This step opens the wood to expose the cellulose for future hydrolysis and converts xylan to xylose.

After pressure letdown in the flash tank, the hydrolyzate is neutralized. After neutralization, the stream flows to the fermentation area. Two percent of the hydrolyzate is sent to cellulase production.

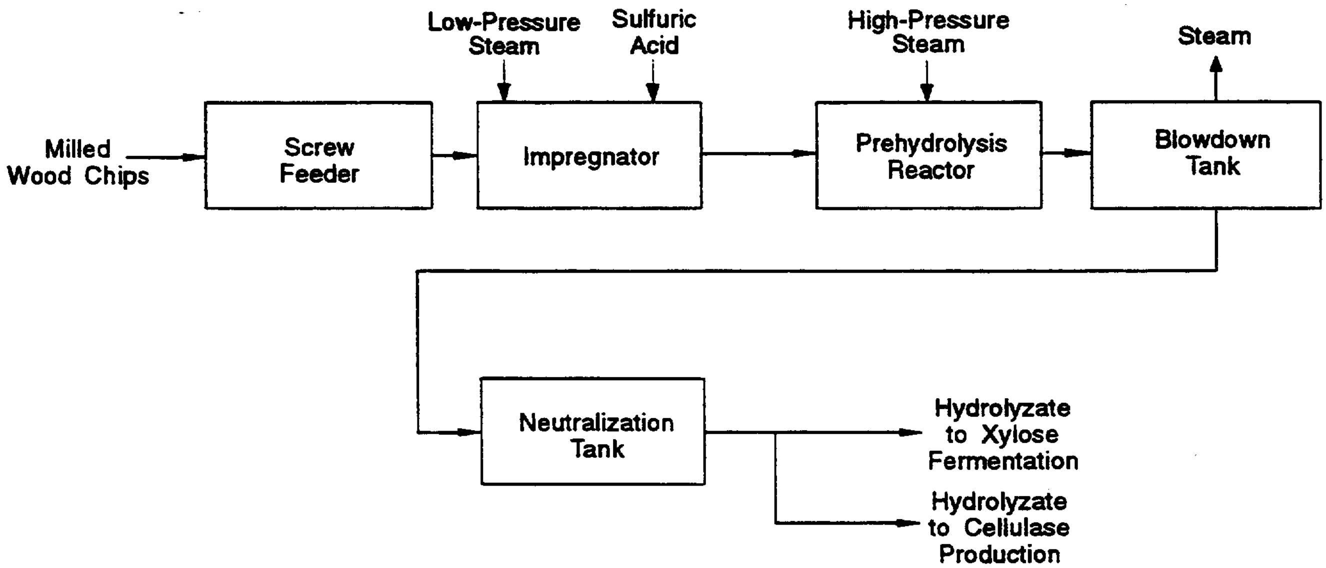
### Comments

Most of the work at NREL on the impregnation and prehydrolysis steps was done on a batch reactor with a standard agitator. Although some work has been done on continuous prehydrolysis, both units should be run continuously at the process conditions to confirm the batch results. Nevertheless, xylose

**Table I-2 — Design Conditions for Prehydrolysis**

| | |
|---|---|
| Temperature | 160 °C |
| Residence Time | 10 minutes |
| Xylan Converted to Xylose | 80% |
| Xylan Converted to Furfural | 13% |
| Xylan Unconverted | 7% |
| Cellulose to Glucose | 3% |
| Cellulose to Hydroxymethyl Furfural (HMF) | 0.1% |
| Cellulose Unconverted | 96.9% |

**Figure I-3 — Section 200: Prehydrolysis and Neutralization**




yields as high as 90 percent have been achieved in the laboratory. While this higher value cannot yet be consistently reproduced, achieving this value on a regular basis is not unreasonable.

Chem Systems has confirmed that the type of prehydrolysis equipment used is based on existing equipment used in the pulp and paper industry and manufactured by companies such as Black Clawson. However, for a plant of this capacity, the vendor recommends two separate lines. This would be feasible, although each proposed reactor would be larger than the current operating equipment. Because of economic considerations associated with minimizing the number of parallel trains and the vendor's recommendation, Chem Systems' cost estimate is based on two prehydrolysis lines.

The material in the neutralization tank following prehydrolysis has a solids content of 12 percent. NREL has mixed and pumped 10 percent material and believes that the 12-percent solids should not present any difficulties. This needs to be confirmed.

## Cellulase Production (Section 300)

### Design Basis

The data are based on batch experiments in the NREL laboratory. The parameters reflect the average performance and are considered by NREL to be conservative. The design basis is shown in Table I-3.

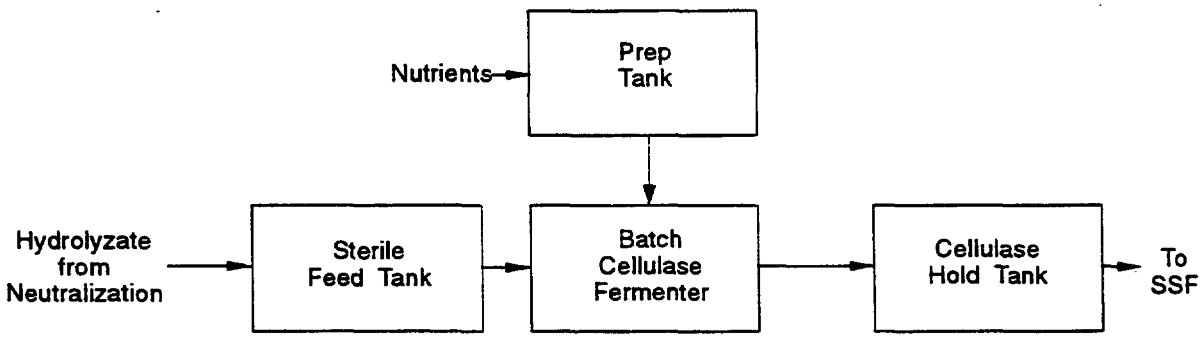
### Process Description

The cellulase production process is shown in Figure I-4. The slipstream of hydrolyzate is fed to a batch fermenter for a 6-day total batch cycle. Seed fermenters feed the main fermenter with cell mass; nutrients and fermentation air enter separately. Chilled water is used to cool the fermenters, which are agitated 250,000-gallon vessels. The cellulase is held in a hold tank and fed continuously to the next step—simultaneous saccharification and fermentation.

### Comments

Although the parameters are an average literature value, the laboratory experiments run by NREL were on ideal substrates. To confirm the design basis, these experiments must be run on actual substrates that contain lignin and other insoluble solids. NREL will attempt to make improvements in the cellulase yield and growth rate.

## Figure I–4 — Section 300: Cellulase Production




### Table I–3 — Design Basis for Cellulase Production

| | |
|---|---|
| Method of Operation | Batch |
| Temperature | 28 °C |
| Pressure | 10 psig |
| Fermentation Time | 5.5 days |
| Cellulase Yield | 202 IU/g cellulose |

## Fermentation (Section 400)

### Design Basis

Simultaneous saccharification and fermentation (SSF) is the key step in NREL's process. The design parameters are based on batch experiments in the NREL laboratory that used lignin and cellulose separated from the liquid after the prehydrolysis step and run on the 50-gallon scale. The solids were reconstituted to the appropriate concentration before being used in the experiment. The design basis is shown in Table I–4.

The data for xylose fermentation are based on NREL laboratory runs in 5-liter reactors using purchased xylose. The system used is based on work done at the University of Florida using E. coli. In addition, Tennessee Valley Authority laboratories ran the xylose fermentation on actual hydrolyzate and found no problems operating. The design basis is shown in Table I–4.

### Process Description

Figure I–5 shows the fermentation process area, which includes xylose fermentation and SSF. The effluent from neutralization flows to the first of 35 fermenters (each with a capacity of 750,000 gallons). Each tank is agitated at very low power (0.1 horsepower per 1,000 gallons). Cell mass is continuously fed into the first SSF fermenter from the SSF seed fermenters. E. coli seed fermenters continuously feed cell mass to the first xylose fermenters. There is no recycle of cell mass in either SSF or xylose fermentation. The effluent from the final fermenter enters the ethanol distillation section.

### Comments

As in cellulase production, the major issue with the NREL lab data is that the experiments did not use actual material that had passed through all the process stages.

NREL claims that the key material that could interfere with the fermentation, lignin, was present in the lab runs; and it believes that gypsum should be inert to the process and have no effect. Nonetheless, running an integrated system in a continuous mode should be a key priority as NREL continues its research.

The yields used are reasonable and reproducible. The major area for process improvement lies in increasing the SSF yield. This is because the yield (72 percent) is relatively low compared to the yields for other process steps and the impact on the cost of production is significant for every percentage-point increase in SSF yield.

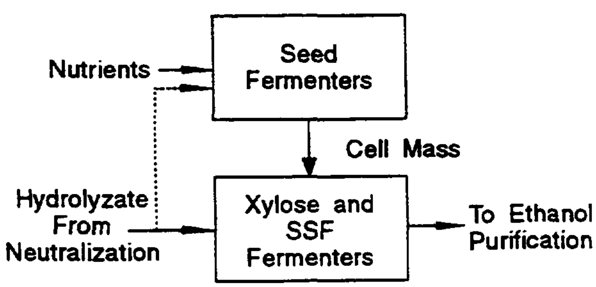
## Table I-4 — Design Basis for Fermentation Area

| | |
|---|---|
| Xylose Fermentation | |
| Xylose Available | 95% |
| Xylose Converted | 90% |
| Fermentation Time | 2 days |
| pH | 7.0 |
| Temperature | 37 °C |
| SSF | |
| Temperature | 37 °C |
| Residence Time | 7 days |
| Cellulose Converted: | |
| to Ethanol and $CO_2$ | 72.0% |
| to Fusel Oils | 0.1% |
| to Glycerol and Acetaldehyde | 4.9% |
| to the Cells | 10.0% |
| Cellulose Unconverted | 13.0% |

The enzyme loading (7 international units per gram of cellulose) is relatively low; a cost analysis should be made to evaluate the relative benefit-to-cost ratio of increasing enzyme loadings (increasing cost) to achieve higher cellulose yields (saving costs).

Inclusion of xylose fermentation is a major advance in the wood-to-ethanol process, increasing the ethanol production by 25 percent over earlier cases. The assumption that 95 percent of the xylose is available is reasonable, and yields are high. There is little room for improvement in the yields because selectivity is nearly 100 percent. An optimum yield may be approximately 90 percent.

## Figure I-5 — Section 400: Simultaneous Saccharification and Fermentation




## Ethanol Purification and Solids Separation (Section 500)

### Design Basis

Because the unit operations are straightforward, NREL has not conducted any lab experiments on this portion of the process. Instead, it is relying on previously engineered systems.

### Process Description

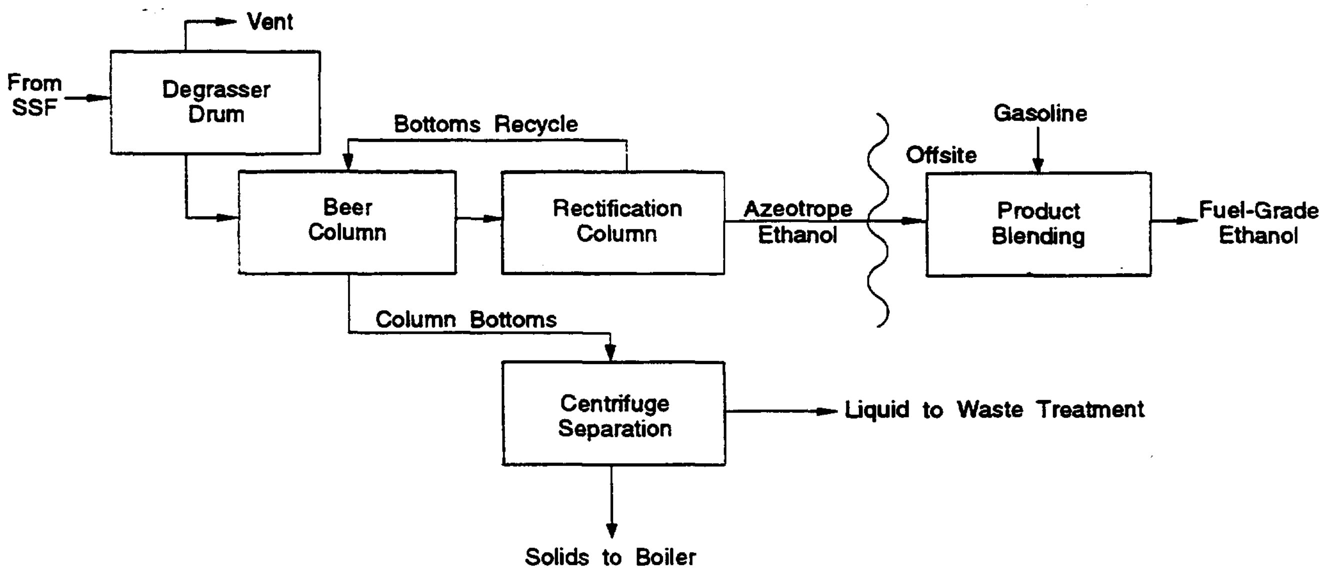
The SSF effluent stream is heated and fed to a degasser drum as depicted in Figure I-6. The carbon dioxide is vented, and the liquid is fed directly into the beer column. This stream contains 1 percent cellulose and 4 percent lignin. The column is operated under slight pressure. The distillate is 40 weight-percent ethanol, 60 weight-percent water. The overheads are fed to the rectification column, where azeotrope ethanol is removed overhead and the bottoms, which contain water and 4 percent ethanol, are recycled to the beer column via the degasser drum. The overhead ethanol-water mixture is mixed with gasoline in the offsite tank area to make the final, fuel-grade product.

The bottoms, containing the suspended and dissolved solids, are fed in parallel to three centrifuges, where use of a supernatant recycle scheme allows recovery of 95 percent of their mass. The solids leaving the centrifuge have a water content of 50 percent. Two screws feed a sludge to a special boiler, where the solids are burned as fuel.

### Comments

The main issue in this process is the nature of the distillation feed—that is, the 5-percent solids content of the feed. A second issue is the low ethanol concentration (4 percent). NREL is basing its design on current practice in corn-to-ethanol plants. According to NREL, the percentage of solids fed to distillation towers operating today is considerably higher and has worse characteristics (that is, they are sticky and more apt to clog the piping) than the finely divided lignin particles. Because there is no commercial operation of a distillation tower with 4 percent lignin in the feed, the viability of this step must be proven on a pilot scale. The alternative scheme, separating the lignin before distillation, might require a more

**Figure I–6 — Section 500: Ethanol Purification and Solids Separation**




complex washing cycle to reduce the amount of ethanol adhering to the solid particles that otherwise would be lost to the product.

The low ethanol concentration in the feed is a result of the solids concentration in the stream leaving the neutralization tank and the yield in the SSF process step. Increasing the solids concentration will increase the final ethanol concentration. Increasing the yield of ethanol during SSF can also increase the ethanol concentration slightly (to the 4.5-percent range). However, this increase in yield will have a positive effect on production economics far more in raw-material costs and capital than in steam cost savings.

The current design basis assumes that the savings in capital costs more than offset higher steam costs incurred by not having process steps to increase ethanol feed concentration. This appears reasonable.

## Waste Treatment (Section 700)

### Design Basis

The design of this section is based on commercially available technology. It contains three process systems: anaerobic digestion, aerobic digestion, and a low-pressure vent system. The liquid from the lignin separation first flows to the anaerobic digester for conversion to methane. Table I–5 contains the design basis for this process.

The methane is sent directly to the boiler as fuel. The liquid from the digester is sent to an aerobic digester. Here, all remaining dissolved solids are assumed to be digested.

### Process Description

Supernatant liquid separated from the lignin and solids in the centrifuge in the separation step after the beer column is sent to waste treatment. Here it enters a holding tank, as shown in Figure I–7, where other streams are mixed before flowing to the anaerobic digester. Ninety percent of the soluble solids, xylose, furfural, and glycerol are converted to methane in this digester. The methane that is produced supplies a substantial amount of the heat released in the boiler. The remaining liquid is

**Table I–5 — Design Basis for Waste Treatment**

| | |
|---|---|
| Conversion of Soluble Solids | 90% |
| Conversion of Xylose | 90% |
| Conversion of Furfural | 90% |
| Conversion of Glycerol | 90% |

sent to an aerobic digester. Lignin, insoluble gypsum, and cell mass are not converted here. From the digester, the liquid is sent to a clarifier, where clear water is separated and discharged. The bottoms of the clarifier are sent to a sludge centrifuge, where they are concentrated to 15 percent solids prior to disposal.

All vents from the plant are fed into a knockout drum with a demister before the vapors are sent to the boiler. The entrained liquid is sent to the anaerobic digester.

**Comments**

The design is based on commercially available technology. However, tests should be made with actual material to confirm yields and throughputs and other characteristics.

## Utilities (Section 800)

Equipment and production rates are discussed in the next chapter.

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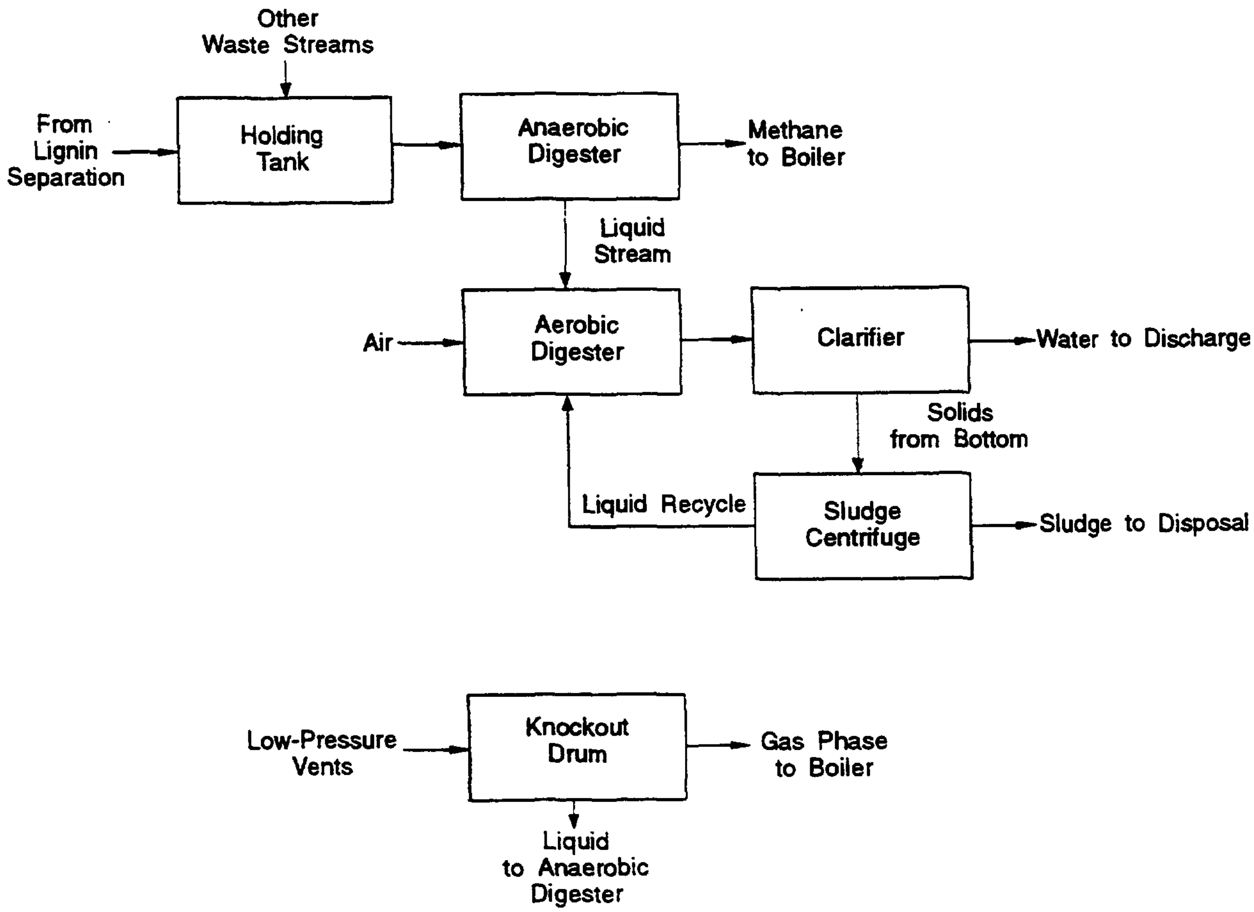
### Figure I-7 — Section 700: Waste Treatment



# II. ECONOMICS

## *BASIS*

Investment and cost-of-production estimates have been developed for a plant producing 57.9 million gallons per year of a denatured ethanol fuel mixture based on a wood feedstock. The fuel mixture is composed of 90.3 weight-percent ethanol, 4.7 weight-percent water, and 5.0 weight-percent gasoline. The plant is based on the NREL process design as described in Chapter I, and the estimate is based on a U.S. first-quarter 1987 timeframe (to be consistent with *Technical Report Three: Methanol Production and Transportation Costs*).

The investment cost has been developed by determining base equipment costs for each piece of equipment. For the major equipment items, which constitute about 80 percent of total equipment costs, prices are based on current vendor estimates. The other items are based on Chem Systems and NREL internal data bases, primarily derived from C COST,® the ICARUS Corporation's cost-estimating computer program, and vendor data.

From the bare equipment cost, installation factors were used to determine the total investment estimate. The installation factors were based on vendor information, specific plant data for fermentation-type plants (as provided to NREL), and Chem Systems' experience.

To be consistent with NREL's format, the estimate of total fixed investment was made for the entire complex without distinguishing between processing units typically considered inside battery limits (ISBL) and auxiliary and supporting facilities typically designated as offsites.

Detailed estimates of utility requirements, capital investment, and production costs are described below.

## *UTILITIES*

### Cogeneration System

The plant is designed with a boiler and power cogeneration system that allows for generation of steam and electricity in a high-pressure steam turbine. The 1,100 pound-per-square-inch absolute (psia) steam boiler is designed to burn gaseous and solid fuels derived from the various organic waste streams in the process. Methane and lignin account for the bulk of the energy value in the fuel stream fed to the boiler.

Gaseous fuels are burned directly. Wet solids are first sent to a drying system that dries and fluidizes the solids into the boiler using boiler flue gas.

The steam and power generating capacities are sized in accordance with the wood-feed rate. The steam turbine is an extracting type that allows for extraction of both 50-pounds-per-square-inch-gauge (psig) and 150-psig steam to meet internal process requirements, with the balance condensed to maximize turbine output.

Based on a steam turbine feed rate of 433,900 pounds per hour of 300 degree Fahrenheit superheated 1,100-psia steam, and extraction of 41,400 pounds per hour of 150-psig steam and 222,900 pounds per hour of 50-psig steam, 36.1 megawatts of power are generated (as calculated by Asea Brown Boveri, a manufacturer of boilers and steam turbines). With total plant demand of 22.8 megawatts, this results in a power surplus of 13.3 megawatts.

### Utility Requirements

Chem Systems has performed a heat balance based on NREL's process design and material balance to determine the utility requirements discussed below.

### Electricity

All of the power requirements in the plant are provided by cogenerated power. Table II–1 contains a summary of electricity requirements by plant section.

The bulk of the electricity requirements are for the disc refiner (Section 100) and the air compressors and refrigeration system (Section 800). The disc refiner requirement is based on 5 horsepower per dry ton per day of wood feed. This is a vendor (Sprout-Bauer) estimate for a similar wood product. However, laboratory testing is needed to confirm the proper morphology of the wood as a feed to prehydrolysis. It is conceivable that the actual power requirement of the refiner could increase substantially based on actual data.

### Steam

The proposed process requires approximately 222,900 pounds per hour of 50-psig steam and 41,400 pounds per hour of 150-psig steam. Table II–2 shows steam requirements by section.

Steam is used primarily in the impregnator and prehydrolysis reactor in the treatment (Section 200) and in the reboiler in the beer column and rectification column in ethanol purification (Section 500).

### Cooling and Process Water

Cooling water (90 degrees Fahrenheit) is available from the cooling tower. A temperature rise of 27 degrees Fahrenheit has been assumed for the process users. Cooling-water requirements are estimated at 19.55 million pounds per hour (39,100 gallons per minute). Table II–3 summarizes cooling-water requirements by section.

Cooling water is primarily required for the steam turbine condenser and the condensers on the ethanol purification columns.

**Table II–2 — Plant Steam Requirements**
(thousand pounds per hour)

| Area No. | Section | 50 psig | 150 psig |
|---|---|---|---|
| 100 | Wood handling | — | — |
| 200 | Pretreatment | 30.6 | 41.4 |
| 300 | Cellulase production | 0.12 | — |
| 400 | Fermentation | — | — |
| 500 | Ethanol purification | 171.1 | — |
| 600 | Offsite tanks | — | — |
| 700 | Waste treatment | — | — |
| 800 | Utilities | 1.1 | — |
| | Miscellaneous | 20.0 | — |
| Total | | 222.92 | 41.4 |

**Table II–1 — Plant Electricity Requirements**
(millions of kilowatthours per year)

| Area No. | Section | Consumption |
|---|---|---|
| 100 | Wood handling | 61.5 |
| 200 | Pretreatment | 3.3 |
| 300 | Cellulase production | 6.0 |
| 400 | Fermentation | 22.7 |
| 500 | Ethanol purification | 3.9 |
| 600 | Offsite tanks | 0.7 |
| 700 | Waste treatment | 3.0 |
| 800 | Utilities | 81.0 |
| Total electricity consumed | | 182.0 |
| Electricity produced | | 288.8 |
| Surplus power produced | | 106.8 |

**Table II–3 — Plant Cooling-Water Requirements**
(gallons per minute)

| Area No. | Section | Consumption |
|---|---|---|
| 100 | Wood handling | 400 |
| 200 | Pretreatment | 7,935 |
| 300 | Cellulase production | — |
| 400 | Fermentation | — |
| 500 | Ethanol purification | 10,400 |
| 600 | Offsite tanks | — |
| 700 | Waste treatment | 3,865 |
| 800 | Utilities | 16,500 |
| Total cooling water consumed | | |
| GPM | | 39,100 |
| Thousand lb/hr | | 19,550 |

In a departure from the original design proposed by NREL, cooling-tower water for fermentations has been replaced by well water. Based on Chem Systems' experience, a temperature rise of 3 degrees Fahrenheit for cooling water (which is necessary in the summer months) for the fermentation units has not been proven feasible in some similar commercial operations.

Using well water (55 degrees Fahrenheit) for xylose fermentation and SSF cooling allows for a 7.5-percent reduction of cooling water flow. The heated well water (assuming a rise of 35 degrees Fahrenheit) can then be utilized for process-water requirements. This scheme allows for better heat integration and results in a slight reduction in capital investment for the cooling-tower system.

Overall, well-water requirements are estimated to be at 2,400 gallons per minute (1.2 million pounds per hour). If well water were not available at a particular site, this requirement would need to be included in the chilled-water system described below.

### Chilled Water

Chilled water (50 degrees Fahrenheit) is provided by a chilled-water system with a refrigeration capacity of 4,000 tons per hour. Table II–4 shows each section's requirements for chilled water.

For chilled water, two different temperature rises are assumed. The lower temperature rise (3.6 degrees Fahrenheit) is used to maximize the condensation of streams leaving the knockout drums in Sections 300 and 400. In cellulase production (Section 300), chilled water with a 27-degree-Fahrenheit temperature rise is used to cool the fermenters. In utilities (Section 800), chilled water is required for interstage cooling of the air compressors.

### Fermentation Air

Process air (45 psig) is required for all the seed fermenters and in cellulase production. The air requirements for these sections are shown in Table II–5.

The NREL design specifies a pressure of 45 psig. Based on vendor information

**Table II–4 — Chilled-Water Requirements**
(gallons per minute)

| | | Consumption @ 50 °F | |
|---|---|---|---|
| Area No. | Section | 3.6 °F delta,T | 27°F delta, T |
| 100 | Wood handling | — | — |
| 200 | Pretreatment | — | — |
| 300 | Cellulase production | 1,840 | 1,435 |
| 400 | Fermentation | 1,895 | — |
| 500 | Ethanol purification | — | — |
| 600 | Offsite tanks | — | — |
| 700 | Waste treatment | — | — |
| 800 | Utilities | — | 1,385 |
| Total chilled water consumed | | | |
| GPM | | 3,735 | 2,820 |
| Thousand lb/hr | | 1,868 | 1,410 |

**Table II–5 — Plant Process Air Requirements**

| Area No. | Section | Lb/Hr |
|---|---|---|
| 100 | Wood handling | — |
| 200 | Pretreatment | — |
| 300 | Cellulase production | 48,400 |
| 400 | Fermentation | 198,100 |
| 500 | Ethanol purification | — |
| 600 | Offsite tanks | — |
| 700 | Waste treatment | — |
| 800 | Utilities | — |
| Total plant air | | |
| Pounds per hour | | 246,500 |
| SCFM | | 56,000 |

(Ingersoll Rand), three compressors would be required to deliver the desired air flow. However, based on information supplied by the vendor, a design that uses 35-psig pressure would allow for a different machine design and the use of a single compressor. Reducing the number of compressors from three to one (excluding spares) would result in a substantial cost savings. Additionally, the vendor believes that some fermentation processes utilize lower pressure air (specifically, 35 psig).

## *CAPITAL SUMMARY*

The breakdown of the total investment cost for the ethanol plant based on the design described above is summarized in Table II–6.

The overall installation factor of 2.85 includes bulk installations, construction labor, site development, buildings, roads, control room, and laboratory, as well as indirect costs, home office costs, engineering, and design. This factor is standard for the type of equipment used. Note that the cost of the steam boiler package, obtained from a vendor estimate, is reported on a separate line item.

Total cost is estimated at $138 million. To be consistent with previous studies, the cost includes items such as owner's costs, land, and startup and commissioning expenses, but does not include financing and any licensing fees.

For this analysis, the capital cost is put on a 1987 basis to be consistent with *Technical Report Three: Methanol Production and Transportation Costs.*

## *PRODUCTION COSTS*

### Base Case

A summary of the cost-of-production estimate for a plant producing about 58 million gallons per year of ethanol-based fuel is presented in Table II–7. The estimate is based on about 1,920 short tons per day (STPD) of dry wood feed. The economics are based on a 1987 U.S. Gulf Coast capital cost estimate. Total capital, as discussed above, is estimated at $138 million.

Cost of production can be divided into several categories:

- **Raw materials**—Primarily wood.
- **Utilities**—Electricity and well water.
- **Operating costs**—Includes labor for operating the plant as well as materials and labor for annual maintenance costs.
- **Overhead expenses**—Includes plant overhead, taxes, and insurance.

Because they are a function of the plant operating rate, raw materials and utilities are considered variable costs. Operating costs and overhead expenses are fixed costs because they are independent of operating rate. The sum of variable and fixed costs is usually termed the cash cost of production. This is the actual out-of-pocket cost an owner incurs before considering depreciation of the capital investment and profits.

A summary of the cost basis is shown in Table II–8, and the detailed cost estimate is presented in Table II–9.

As in earlier reports evaluating the conversion of natural gas and coal to methanol and biomass gasification to methanol, a capital charge of 20 percent of the total fixed investment plus working capital is taken as an overall capital-recovery factor. This is equivalent to a discounted cash flow after-tax rate of return of approximately 10 percent.

In the ethanol cost estimates, raw materials—the largest component of the production cost—are estimated at a net cost of $0.60 per gallon of fuel. The major component is wood, taken at $42 per short ton on a dry basis.

Because the plant has a cogeneration system that uses waste materials as fuel to the boiler, the plant is a net producer of power. The only external utility is well water.

Based on a net utility credit of $0.054 per gallon of fuel product, the total variable cost of the material is estimated at $0.55 per gallon. The actual power consumption in the woodmill section is subject to actual vendor testing verification.

The unit cost of exported electricity is taken at $0.03 per kilowatthour. This is the rate at which it is assumed the ethanol plant can sell its excess electricity to a utility. This rate is consistent with rates at which cogeneration plants currently sell to utilities.

Direct cash cost, including labor, maintenance, and direct plant overheads, totals about $0.11 per gallon of product. Operating labor is based on nine workers per shift: one in the control room, two in wood handling, two in the process, one in the tank farm and blending, one for waste handling, and two for utilities.

## Table II-6 — Capital Cost Summary

| Section/Item | Purchased Cost ($MM/Unit) 1990 | No. Req'd. | Total Purchased ($MM) $1990 | $1987 | Source[a] |
|---|---|---|---|---|---|
| Wood Handling (Section 100) | | | | | |
| Major Equipment | | | | | |
| Disc Refiners | 0.37 | 4 | 1.48 | | Sprout-Bauer |
| Front End Loaders | 0.16 | 3 | 0.47 | | |
| Belt Conveyor | 0.19 | 1 | 0.19 | | |
| Other | | | 0.36 | | |
| Section Total | | | 2.50 | | |
| Prehydrolysis (Section 200) | | | | | |
| Major Equipment | | | | | |
| Impregnator/Prehydrolysis System | 3.66 | 2 | 7.32 | | Black Clawson |
| Screwfeeder | 0.28 | 2 | 0.55 | | Black Clawson |
| Other | | | 0.28 | | |
| Section Total | | | 8.15 | | |
| Cellulase Production (Section 300) | | | | | |
| Major Equipment | | | | | |
| Cellulase Fermenter | 0.07 | 3 | 0.20 | | |
| Fermenter Agitator | 0.08 | 3 | 0.23 | | |
| Feed Tank Agitator | 0.13 | 1 | 0.13 | | |
| Other | | | 0.39 | | |
| Section Total | | | 0.94 | | |
| Fermentation (Section 400) | | | | | |
| Major Equipment | | | | | |
| SSF Fermenter | 0.20 | 35 | 7.00 | | CBI |
| SSF Fermenter | 0.31 | 1 | 0.31 | | CBI |
| Seed Hold Tank | 0.12 | 1 | 0.32 | | CBI |
| Seed Hold Tank | 0.19 | 1 | 0.41 | | CBI |
| SSF Fermenter Agitators | 0.03 | 27 | 0.81 | | |
| Other | | | 1.14 | | |
| Section Total | | | 9.99 | | |
| Distillation (Section 500) | | | | | |
| Major Equipment | | | | | |
| Centrifuge | 0.23 | 3 | 0.68 | | |
| Beer Column | 0.17 | 1 | 0.17 | | |
| Rectification Column | 0.16 | 1 | 0.16 | | |
| Other | | | 0.36 | | |
| Section Total | | | 1.36 | | |

## Table II-6 — Capital Cost Summary (continued)

| Section/Item | Purchased Cost ($MM/Unit) 1990 | No. Req'd. | Total Purchased ($MM) $1990 | $1987 | Source[a] |
|---|---|---|---|---|---|
| **Offsite Tankage (Section 600)** | | | | | |
| Major Equipment | | | | | |
| Ethanol Product Tank | 0.25 | 2 | 0.49 | | CBI |
| $NH_3$ Storage Tank | 0.09 | 2 | 0.17 | | |
| Fire Water Tank | 0.14 | 1 | 0.14 | | |
| Other | | | 0.30 | | |
| Section Total | | | 1.10 | | |
| **Environmental and Wastewater (Section 700)** | | | | | |
| Major Equipment | | | | | |
| Secondary Clarifier | 0.26 | 1 | 0.26 | | |
| LP Vent Blower | 0.07 | 2 | 0.15 | | |
| Equalization Tank | 0.24 | 1 | 0.24 | | |
| Other | | | 0.80 | | |
| Section Total | | | 1.44 | | |
| **Boiler and Steam Distribution (Section 800)** | | | | | |
| Major Equipment | | | | | |
| Cooling Tower System | 0.73 | 1 | 0.73 | | |
| Demineralizers | 0.31 | 2 | 0.62 | | |
| Condensate Polisher | 0.10 | 2 | 0.20 | | |
| Turbo Generator | 6.50 | 1 | 6.50 | | ABB |
| Air Compressor Package | 0.45 | 4 | 1.80 | | Ingersoll-Rand |
| Chilled-Water Package | 0.60 | 1 | 0.60 | | |
| Other | | | 1.20 | | Trane |
| Section Total | | | 11.65 | | |
| Plant Subtotal | | | 37.12 | 33.7 | |
| Miscellaneous Equipment | | | 2.23 | 2.02 | |
| Total—Bare Equipment | | | 39.35 | 35.72 (4%) | |
| Total Installed Cost-Factor = 2.85 | | | 112.14 | | |
| Steam Boiler Package (w/predryer) | | | installed 19.80 | 18 (6.32) | ABB |
| Total Plant Investment | | | 131.94 | 120.07 | |
| Owner's Cost, Fees, and Profit | | | 13.19 | 12.01 | |
| Startup Cost | | | | 6.00 | |
| Grand Total Plant Investment | | | | 138.07 | |

[a]On a purchased equipment basis, about 80 percent of the equipment is based on current vendor budgetary estimates.

## Table II-7 — Ethanol-Based Fuel Economic Summary (1987 - U.S. Gulf Coast)

| | |
|---|---|
| Investment | $138 MM |
| Working capital | 9 MM |
| Production cost, $/gal | |
| Net raw materials | 0.60 |
| Utilities | (0.05) |
| Direct cash cost | 0.11 |
| Allocated cash cost | 0.10 |
| Full cash cost | 0.76 |
| Cost plus 20% capital charges | $1.27 |

Allocated cash costs, which include general plant overhead and insurance and local property taxes, contribute a total of $0.10 to the cost of production. This results in a total cash cost of $0.76 per gallon of fuel ethanol.

Adding a 20-percent capital-recovery charge, reflecting both depreciation and return on investment, the required ethanol fuel price would be $1.27 per gallon of fuel. This capital-recovery rate is equivalent to a discounted cash-flow after-tax rate of return of approximately 10 percent and is consistent with the treatment of capital costs used in Technical Reports Three and Five.

This production cost estimate does not include items such as shipping, packaging, research and development expenses, general sales and administrative costs, and royalties.

### Sensitivities

The base-case economics have been based on an evaluation of NREL experimental data. Accordingly, a number of cost sensitivities have been carried out to illustrate the effect of various parameters.

### Capital Sensitivity

Investment cost has been based on a factored estimate and may vary, depending on the design philosophy used in the plant. This type

## Table II-8 — Bases for Ethanol Production Costs

- 4th quarter, 1987.
- Operating factor: 91 percent, 8,000 hours per year.
- Direct overhead at 45 percent of labor and supervision.
- General plant overhead at 65 percent of operating costs.
- Maintenance at 3 percent of total fixed investment.
- Insurance and property taxes at 1.5 percent of total fixed investment.
- Working capital is recovered at the end of the life of the project and is calculated as the sum of the following four items:

  1. **Accounts receivable**—1.00 month's gross cost of production (COP)[a]
  2. **Cash**—1.00 week's gross COP - depreciation[a]
  3. **Warehouse/spares**—3.00 percent inside battery limits (ISBL)
  4. **Accounts payable**—1.00 month's raw materials

- Capital charges at 20 percent of total capital requirements (fixed plus working capital). This charge covers depreciation recovery and a return in capital.

[a]Gross COP = net COP (excluding increases in working capital) less byproduct credit.

## Table II–9 — Cost-of-Production Estimate for Denatured Fuel (90.25% Ethanol) From NREL Wood-to-Ethanol Process (68% Overall Yield)

| | | | Capital Cost ($MM) | | |
|---|---|---|---|---|---|
| | | | Orig. | Book | Repl. |
| Plant Startup: | 1987 | | | | |
| Analysis: | Fourth quarter, 1987 | | | | |
| Location: | U.S. | | | | |
| Capacity: | 57.91 million gallons per year<br>173,378 metric tons per year | | | | |
| Onstream time: | 8,000 hours per year | Total fixed inv. | 138.1 | 138.1 | 138.1 |
| Throughput: | 57.91 million gallons per year | Working capital | | | 9.1 |

### Production Cost Summary

| Component | | Units per gal | Price ($/unit) | $ per gal[a] | Annual Cost ($MM) | $ per met. ton |
|---|---|---|---|---|---|---|
| Raw materials | Wood (dry), ST | 0.0111 | 42.000 | 0.464 | 26.88 | |
| | Sulfuric acid, lb | 0.3976 | 0.032 | 0.013 | 0.74 | |
| | Lime, lb | 0.2940 | 0.023 | 0.007 | 0.38 | |
| | Ammonia, lb | 0.6296 | 0.041 | 0.026 | 1.50 | |
| | Nutrients, lb | 0.0181 | 0.115 | 0.002 | 0.12 | |
| | Corn steep liquor, lb | 0.0633 | 0.100 | 0.006 | 0.37 | |
| | Corn oil (antifoam), lb | 0.0039 | 0.240 | 0.001 | 0.05 | |
| | Glucose, lb | 0.0496 | 0.510 | 0.025 | 1.46 | |
| | Gasoline/diesel, gal | 0.0570 | 0.770 | 0.044 | 2.54 | |
| | Catalyst & chemicals | | 0.010 | 0.010 | 0.58 | |
| | Total raw materials | | | 0.598 | 34.62 | 200 |
| Byproduct credits | Solids disposal, ton | (0.00034) | 20.000 | 0.007 | 0.40 | |
| | Total byproduct credits | | | 0.007 | 0.40 | 2 |
| Net raw materials | | | | 0.605 | 35.02 | 202 |
| Utilities | Power, kWh | (1.85300) | 0.030 | (0.056) | (3.22) | |
| | Well water, M gal | 0.01987 | 0.100 | 0.002 | 0.12 | |
| | Total Utilities | | | (0.054) | (3.10) | (18) |
| Variable cost of production | | | | 0.551 | 31.91 | 184 |
| Direct cash costs | Labor (41 men @ $29,800/yr) | | | 0.021 | 1.22 | |
| | Foremen (9 men @ $34,000/yr) | | | 0.005 | 0.31 | |
| | Supervisors (1 man @ $40,000/yr) | | | 0.001 | 0.04 | |
| | Maintenance, material, & labor (3% of ISBL) | | | 0.072 | 4.14 | |
| | Direct overhead (45% labor/supervision) | | | 0.012 | 0.71 | |
| | Total direct cash costs | | | 0.111 | 6.42 | 37 |
| Allocated cash costs | General plant overhead (65% labor/maintenance) | | | 0.064 | 3.71 | |
| | Insurance, property tax (1.5% total fixed inv.) | | | 0.036 | 2.07 | |
| | Total allocated cash costs | | | 0.100 | 5.78 | 33 |
| Full cash cost of production | | | | 0.762 | 44.11 | 254 |
| Net cost of production | | | | 0.762 | 44.11 | 254 |
| Cost plus 20% capital charges | | | | 1.270 | 73.55 | 424 |

[a]Gallon of denatured, hydrous ethanol.

of budgetary estimate is considered accurate plus or minus 30 percent. Figure II–1 illustrates the effect of ethanol production economics on investment cost.

**Wood Cost**

The wood-feed system is a complex system and a major component of the plant's operational feasibility. Accordingly, the actual cost of wood for an operation of this size could vary from the base estimate. Figure II–2 shows ethanol fuel production cost as a function of the cost of dry wood.

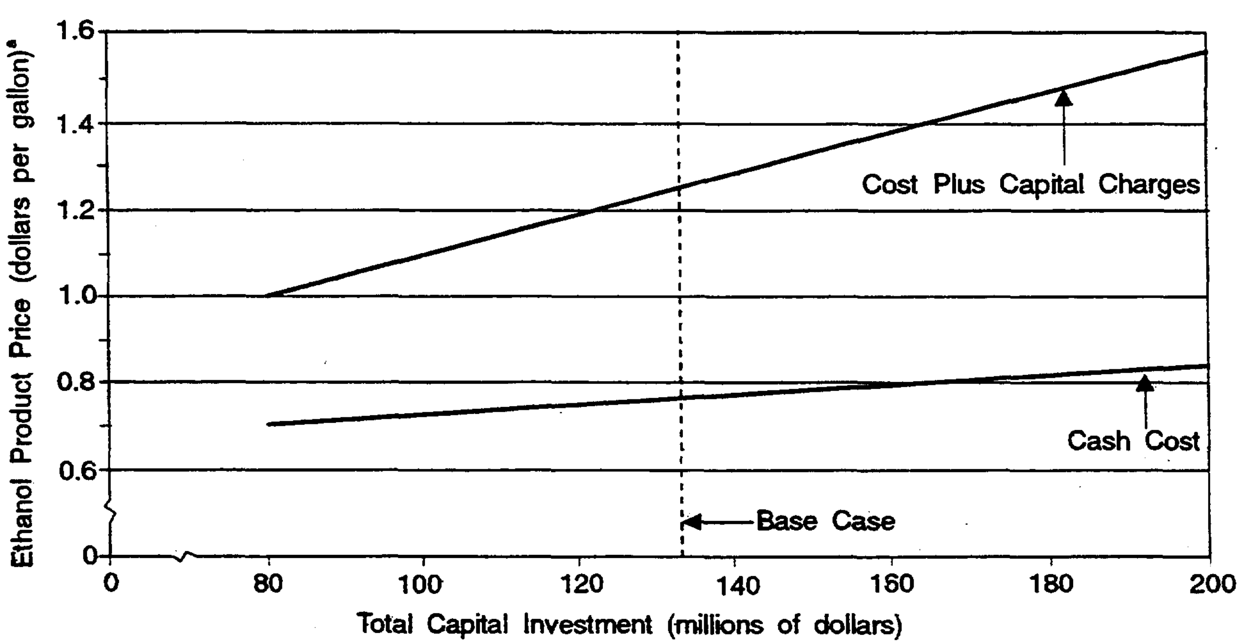
**Yield**

The overall yield for the base case is 68 percent, based on carbohydrates (hemicellulose and cellulose). If improvements can be made in the various processing steps, the overall yield can increase substantially. NREL believes that there is a strong basis for improved yields and cites the following points:

- Xylan-to-xylose yields in the laboratory have been as high as 90 percent, which proves that such yields can be achieved. The base case uses 80 percent.
- Xylose-to-ethanol conversion using existing better organisms that have been tested in other laboratories can boost yield to between 90 and 95 percent instead of the 85.5 percent assumed in the base case.
- The amount of cellulose used for cellulase production and seed fermenters can be reduced.
- NREL claims that experimental evidence indicates that ethanol yield from cellulose during SSF can reach as high as 95 percent instead of the yield of 72 percent assumed in the base case.

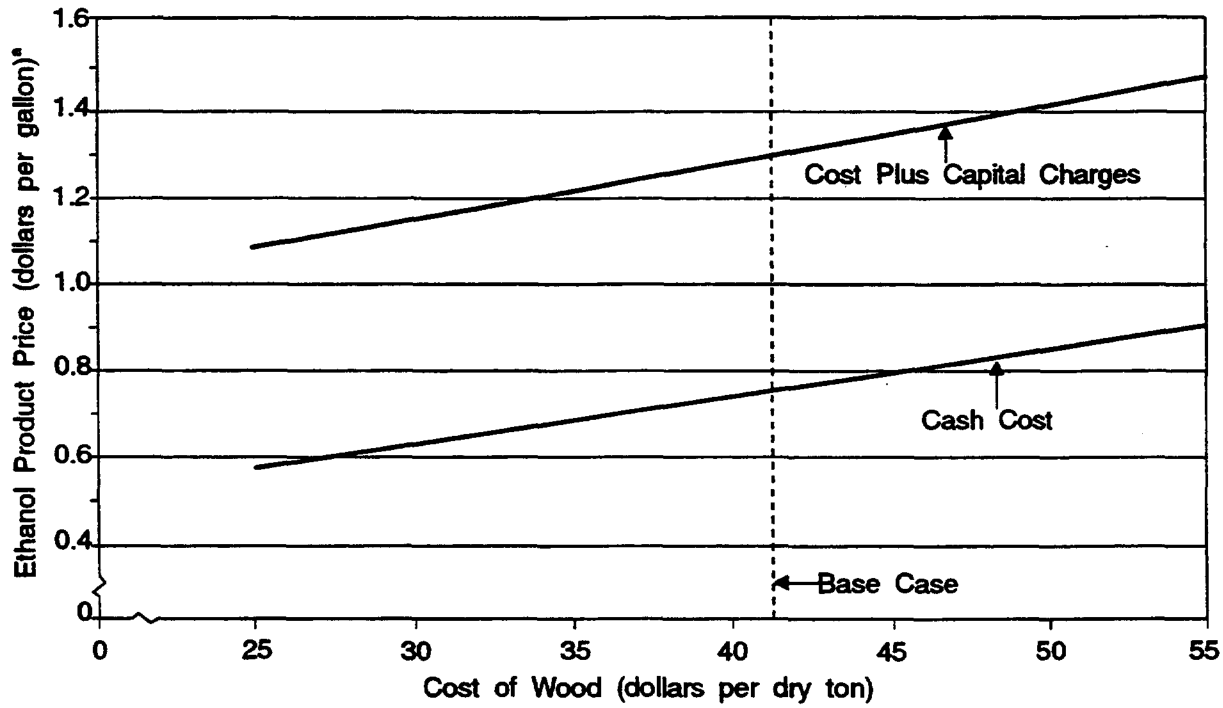
Estimates have been made of the effect of increased yield on ethanol cost at a constant wood-feed rate, resulting in increased capacity at the same capital cost. Table II–10 presents a cost-of-production estimate for a case at 90 percent overall carbohydrate yield. It should be pointed out that as the yield increases, the amount of carbohydrate not converted to

**Figure II–1 — Effect of Capital Investment Cost on Ethanol Product Price, 1,920 STPD Dry Wood Feed**




*Last quarter of 1987.

## Figure II-2 — Effect of Wood Cost on Ethanol Product Price, 1,920 STPD Dry Wood Feed




*Last quarter of 1987.

ethanol decreases; thus the amount available for use as a boiler fuel also is reduced. Therefore, the improved yield is accompanied with a reduction, albeit small, in the amount of electricity exported.

The effect of overall yield on ethanol prices is shown in Figure II–3. Capacity for the base case at various yields increases and is listed in Table II–11.

As mentioned above, all yields and potential improvements need to be verified on actual runs using material that flows from one process section to another. This is essential for analysis of the effect of byproducts formed during upstream steps and carried forward to subsequent steps throughout the process.

### Plant Size

To evaluate the effects of increased capacity on ethanol price, an analysis was made for a plant having a capacity of 5 times the base case, or 9,600 short tons per day of dry wood feed. This case was chosen to match previous evaluations of production of methanol from natural gas, coal, and biomass. There is a need, of course, to verify the costs and logistics of wood collection, delivery, and renewal for such a large plant. This verification is beyond the scope of this study.

Assuming the viability of a plant that can process 10,000 tons per day of dry wood feed, a cost of production estimate was prepared at various overall yields. Tables II–12 and II–13 present the economics for such a plant at the base-case yield and 90 percent overall yield. Figure II–4 illustrates the effect of overall yield on ethanol product price for this large-capacity case.

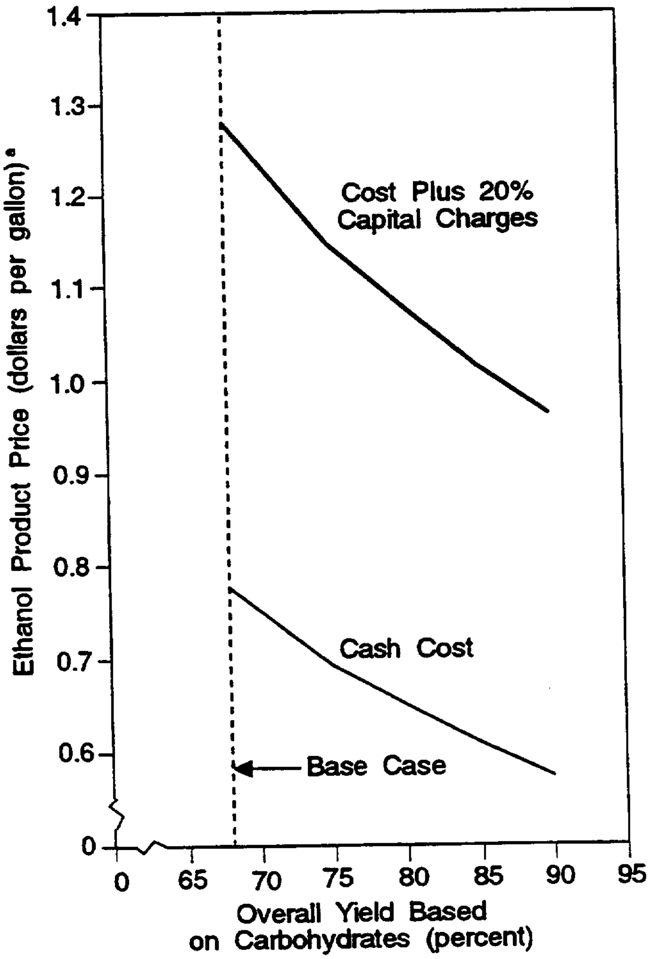
## Table II-10 — Cost-of-Production Estimate for Denatured Fuel (90.25% Ethanol) From NREL Wood-to-Ethanol Process (90% Overall Yield)

| | | | Capital Cost ($MM) | | |
|---|---|---|---|---|---|
| | | | Orig. | Book | Repl. |
| Plant startup: | 1987 | | | | |
| Analysis: | Fourth quarter, 1987 | Battery limits | 138.1 | 138.1 | 138.1 |
| Location: | U.S. | Offsites | 0.0 | 0.0 | 0.0 |
| Capacity: | 77.50 million gallons per year | | | | |
| | 232,015 metric tons per year | | | | |
| Onstream time: | 8,000 hours per year | Total Fixed Inv. | 138.1 | 138.1 | 138.1 |
| Throughput: | 77.50 million gallons per year | Working Capital | | | 9.2 |

### Production Cost Summary

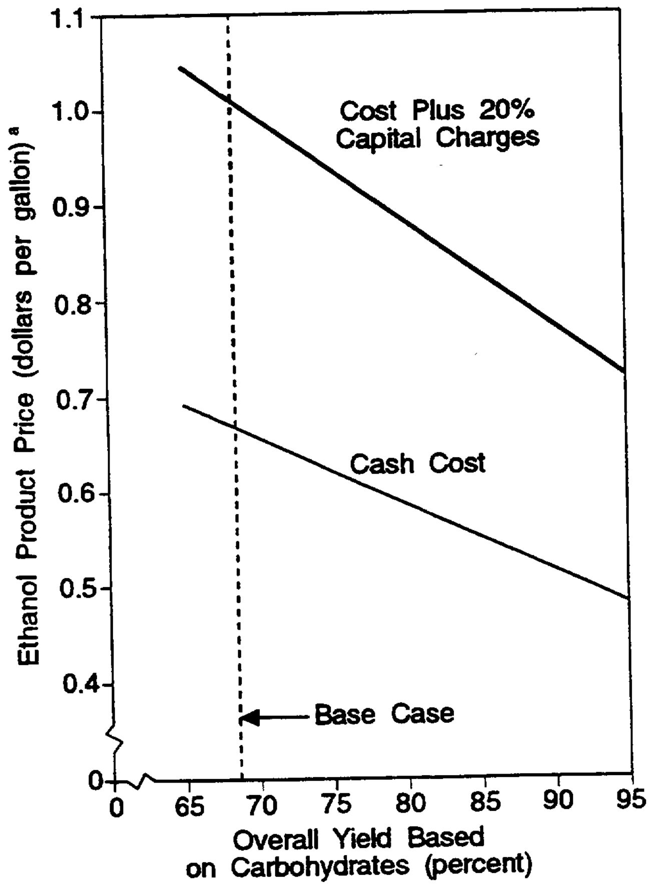
| Component | | Units per gal | Price ($/unit) | $ per gal | Annual Cost ($MM) | $ per met. ton |
|---|---|---|---|---|---|---|
| Raw materials | Wood (dry), ST | 0.0083 | 42.000 | 0.347 | 26.88 | |
| | Sulfuric acid, lb | 0.2970 | 0.032 | 0.010 | 0.74 | |
| | Lime, lb | 0.2190 | 0.023 | 0.005 | 0.38 | |
| | Ammonia, lb | 0.4704 | 0.041 | 0.019 | 1.49 | |
| | Nutrients, lb | 0.0181 | 0.115 | 0.002 | 0.16 | |
| | Corn steep liquor, lb | 0.0633 | 0.100 | 0.006 | 0.49 | |
| | Corn oil (antifoam), lb | 0.0039 | 0.240 | 0.001 | 0.07 | |
| | Glucose, lb | 0.0370 | 0.510 | 0.019 | 1.46 | |
| | Gasoline/diesel, gal | 0.0570 | 0.770 | 0.044 | 3.40 | |
| | Catalyst & chemicals | | 0.010 | 0.010 | 0.78 | |
| | Total raw materials | | | 0.463 | 35.85 | 155 |
| Byproduct credits | Solids disposal, ton | (0.00034) | 20.000 | 0.007 | 0.53 | |
| | Total byproduct credits | | | 0.007 | 0.53 | 2 |
| Net raw materials | | | 0.470 | 36.39 | 157 | |
| Utilities | Power, kWh | (1.48400) | 0.030 | (0.045) | (3.45) | |
| | Well water, M gal | 0.01987 | 0.100 | 0.002 | 0.15 | |
| | Total Utilities | | | (0.043) | (3.30) | (14) |
| Variable cost of production | | | | 0.427 | 33.09 | 143 |
| Direct cash costs | Labor (41 men @ $29,800/yr) | | | 0.016 | 1.22 | |
| | Foremen (9 men @ $34,000/yr) | | | 0.004 | 0.31 | |
| | Supervisor (1 man @ $40,000/yr) | | | 0.001 | 0.04 | |
| | Maintenance, material, & labor (3% of ISBL) | | | 0.053 | 4.14 | |
| | Direct overhead (45% labor/supervision) | | | 0.009 | 0.71 | |
| | Total direct cash costs | | | 0.083 | 6.42 | 28 |
| Allocated Cash costs | General plant overhead (65% labor/maintenance) | | | 0.048 | 3.71 | |
| | Insurance, property tax (1.5% total fixed inv.) | | | 0.027 | 2.07 | |
| | Total allocated cash costs | | | 0.75 | 5.78 | 25 |
| Full cash cost of production | | | | 0.584 | 45.29 | 195 |
| Net cost of production | | | | 0.584 | 45.29 | 195 |
| Cost plus 0% return on total book investment plus working capital | | | | 0.584 | 45.29 | 195 |
| Cost plus 20% return on total book investment plus working capital | | | | 0.965 | 74.75 | 322 |
| Cost plus 30% return on total book investment plus working capital | | | | 1.155 | 89.48 | 386 |

**Figure II-3 — Effect of Improved Overall Yield on Ethanol Product Price, 1,920 STPD Dry Wood Feed**




[a] Last quarter of 1987.

**Figure II-4 — Effect of Overall Yield on Ethanol Product Price for a Large Capacity Plant, 9,600 STPD Dry Wood Feed**




[a] Last quarter of 1987.

**Table II-11 — Effect of Overall Yield on Ethanol Prices**

| Yield (%) | Capacity (millions of gal/yr) |
|---|---|
| 68 | 57.9 |
| 75 | 64.6 |
| 85 | 73.2 |
| 90 | 77.5 |

## Table II-12 — Cost-of-Production Estimate for Denatured Fuel (90.25% Ethanol) From NREL Wood-to-Ethanol Process at Large Plant (68% Overall Yield)

| | | | Capital Cost ($MM) | | |
|---|---|---|---|---|---|
| | | | Orig. | Book | Repl. |
| Plant Start-up: | 1987 | | | | |
| Analysis Date: | Fourth quarter, 1987 | | | | |
| Location: | U.S. | Battery limits | 466.7 | 466.7 | 466.7 |
| Capacity: | 289.55 million gallons per year | Offsites | 0.0 | 0.0 | 0.0 |
| | 866,838 metric tons per year | | | | |
| Onstream time: | 8,000 hours per year | Total Fixed Inv. | 466.7 | 466.7 | 466.7 |
| Throughput: | 289.55 million gallons per year | Working Capital | | | 35.2 |

### Production Cost Summary

| Component | | Units per gal | Price ($/unit) | $ per gal | Annual Cost ($MM) | $ per met. ton |
|---|---|---|---|---|---|---|
| Raw materials | Wood (dry), ST | 0.0111 | 42.000 | 0.464 | 134.39 | |
| | Sulfuric acid, lb | 0.3976 | 0.032 | 0.013 | 3.68 | |
| | Lime, lb | 0.2940 | 0.023 | 0.007 | 1.92 | |
| | Ammonia, lb | 0.6296 | 0.041 | 0.026 | 7.47 | |
| | Nutrients, lb | 0.0181 | 0.115 | 0.002 | 0.60 | |
| | Corn steep liquor, lb | 0.0633 | 0.100 | 0.006 | 1.83 | |
| | Corn oil (antifoam), lb | 0.0039 | 0.240 | 0.001 | 0.27 | |
| | Glucose, lb | 0.0496 | 0.510 | 0.025 | 7.32 | |
| | Gasoline/diesel, gal | 0.0570 | 0.770 | 0.044 | 12.70 | |
| | Catalyst & chemicals | | 0.010 | 0.010 | 2.90 | |
| | Total raw materials | | | 0.598 | 173.19 | 200 |
| Byproduct credits | Solids disposal, ton | (0.00034) | 20.000 | 0.007 | 2.00 | |
| | Total byproduct credits | | | 0.007 | 2.00 | 2 |
| Net raw materials | | | 0.605 | 175.09 | 202 | |
| Utilities | Power, kWh | (1.85300) | 0.030 | (0.056) | (16.10) | |
| | Well water, M gal | 0.01987 | 0.100 | 0.002 | 0.58 | |
| | Total Utilities | | | (0.054) | (15.52) | (18) |
| Variable cost of production | | | | 0.551 | 159.57 | 184 |
| Direct cash costs | Labor (41 men @ $29,800/yr) | | | 0.008 | 2.44 | |
| | Foremen (9 men @ $34,000/yr) | | | 0.002 | 0.61 | |
| | Supervisor (1 man @ $40,000) | | | 0.000 | 0.08 | |
| | Maintenance, material, & labor (3% of ISBL) | | | 0.048 | 14.00 | |
| | Direct overhead (45% labor/supervision) | | | 0.005 | 1.41 | |
| | Total direct cash costs | | | 0.064 | 18.55 | 21 |
| Allocated cash costs | General plant overhead (65% labor/maintenance) | | | 0.038 | 11.14 | |
| | Insurance, property tax (1.5% total fixed inv.) | | | 0.024 | 7.00 | |
| | Total allocated cash costs | | | 0.063 | 18.14 | 21 |
| Full cash cost of production | | | | 0.678 | 196.25 | 226 |
| Net cost of production | | | | 0.678 | 196.25 | 226 |
| Cost plus 0% return on total book investment plus working capital | | | | 0.678 | 196.25 | 226 |
| Cost plus 20% capital charges | | | | 1.024 | 296.64 | 342 |
| Cost plus 30% return on total book investment plus working capital | | | | 1.198 | 346.83 | 400 |

## Table II-13 — Cost-of-Production Estimate for Denatured Fuel (90.25% Ethanol) From NREL Wood-to-Ethanol Large Plant Process (90% Overall Yield)

| | |
|---|---|
| Plant Startup: | 1987 |
| Analysis: | Fourth quarter, 1987 |
| Location: | U.S. |
| Capacity: | 387.50 million gallons per year<br>1,160 metric tons per year |
| Onstream time: | 8,000 hours per year |
| Throughput: | 387.50 million gallons per year |

| | Capital Cost ($MM) | | |
|---|---|---|---|
| | Orig. | Book | Repl. |
| Battery limits | 466.7 | 466.7 | 466.7 |
| Offsites | 0.0 | 0.0 | 0.0 |
| Total fixed inv. | 466.7 | 466.7 | 466.7 |
| Working capital | | | 35.8 |

### Production Cost Summary

| | | Units per gal | Price ($/unit) | $ per gal | Annual Cost ($MM) | $ per met. ton |
|---|---|---|---|---|---|---|
| Raw materials | Wood (dry), ST | 0.0083 | 42.000 | 0.347 | 134.40 | |
| | Sulfuric acid, lb | 0.2970 | 0.032 | 0.010 | 3.68 | |
| | Lime, lb | 0.2190 | 0.023 | 0.005 | 1.91 | |
| | Ammonia, lb | 0.4704 | 0.041 | 0.019 | 7.47 | |
| | Nutrients, lb | 0.0181 | 0.115 | 0.002 | 0.81 | |
| | Corn steep liquor, lb | 0.0633 | 0.100 | 0.006 | 2.45 | |
| | Corn oil (antifoam), lb | 0.0039 | 0.240 | 0.001 | 0.36 | |
| | Glucose, lb | 0.0370 | 0.510 | 0.019 | 7.31 | |
| | Gasoline/diesel, gal | 0.0570 | 0.770 | 0.044 | 17.00 | |
| | Catalyst & chemicals | | 0.010 | 0.010 | 3.88 | |
| | Total raw materials | | | 0.463 | 179.27 | 155 |
| Byproduct credits | Solids disposal, ton | (0.00034) | 20.000 | 0.007 | 2.67 | |
| | Total byproduct credits | | | 0.007 | 2.67 | 2 |
| Net raw materials | | | 0.470 | 181.94 | 157 | |
| Utilities | Power, kWh | (1.48400) | 0.030 | (0.045) | (17.25) | |
| | Well water, M gal | 0.01987 | 0.100 | 0.002 | 0.77 | |
| | Total utilities | | | (0.043) | (16.48) | (14) |
| Variable cost of production | | | | 0.427 | 165.46 | 143 |
| Direct cash costs | Labor (41 men @ $29,800/yr) | | | 0.006 | 2.44 | |
| | Foremen (9 men @ $34,000/yr) | | | 0.002 | 0.61 | |
| | Supervisors (1 man @ $40,000) | | | 0.000 | 0.08 | |
| | Maintenance, material, & labor (3% of ISBL) | | | 0.036 | 14.00 | |
| | Direct overhead (45% labor/supervision) | | | 0.004 | 1.41 | |
| | Total direct cash costs | | | 0.048 | 18.55 | 16 |
| Allocated cash costs | General plant overhead (65% labor/maintenance) | | | 0.029 | 11.14 | |
| | Insurance, property tax (1.5% total fixed inv.) | | | 0.018 | 7.00 | |
| | Total allocated cash costs | | | 0.047 | 18.14 | 16 |
| Full cash cost of production | | | | 0.522 | 202.15 | 174 |
| Net cost of production | | | | 0.522 | 202.15 | 174 |
| Cost plus 0% return on total book investment plus working capital | | | | 0.522 | 202.15 | 174 |
| Cost plus 20% return on total book investment plus working capital | | | | 0.781 | 302.65 | 261 |
| Cost plus 30% return on total book investment plus working capital | | | | 0.911 | 352.90 | 304 |

# APPENDIX

## *STEAM BOILER*

Type of quote: Verbal.

Contact: Andy Sefcik, A.B.B. (201) 992-2392.

Conditions:

## *BOILER FUEL*

See Table A–1.

**Table A–1 — Boiler Fuel**

| Boiler Fuel | Btu/Lb (LVH) | Lb/Hr | Btu/Hr |
|---|---|---|---|
| Water | 0 | 83,963 | 0 |
| Cellulose | 6,960 | 8,588 | 59,772,480 |
| Xylan | 6,510 | 525 | 3,417,750 |
| Soluble solids | — | 1,093 | 0 |
| Ash | — | 39 | 0 |
| Lignin | 10,650 | 38,401 | 408,970,650 |
| Xylose | 6,510 | 513 | 3,339,630 |
| HMF | 6,510 | 7 | 45,570 |
| Gypsum (soluble) | — | 1,189 | 0 |
| Gypsum (insoluble) | — | 4,765 | 0 |
| $CO_2$ | 0 | 6,797 | 0 |
| Cellulose | 6,960 | 44 | 306,240 |
| Glycerol | — | 231 | 0 |
| Cell mass | 5,000 | 9,669 | 48,345,000 |
| Methane | 21,500 | 5,538 | 119,067,000 |
| Total | | 160,362 | 643,264,320 |

### Exit Pressure

Exit pressure: 1,100 psia @ 857 °F (300 °F of superheat).

Cost (1990): $19,800,000 installed.

Note: A.B.B. subsidiaries include the company that produces Flakt-type dryers. Therefore, quote encompasses same type of design as in the earlier Badger report.

## *STEAM TURBINE*

Type of quote: Verbal.

Contact: Bill Krohner, A.B.B. (203) 673-7463.

Conditions:

Steam turbine feed: 433,878/hr of steam, 1,100 psia with 300 °F of superheat (875 °F).

Extraction @ 150 psig: 41,354 lb/hr

Extraction @ 50 psig: 222,853 lb/hr

Steam turbine exit pressure: 89 MM Hg

Cooling water @ 90 °F

Chilled water @ 50 °F

Cost (1990): $6,500,000 bare equipment. Includes turbine and condenser. Not included are foundation, erection and supervision, electrical package, piping, extraction, and expansion valves.

Note: According to Mr. Krohner, given the above conditions, a turbine of this type should be able to produce approximately 36 MW of power.

## *DISK REFINER*

Type of quote: Written, nonbinding.

Contact: David Kenamond, A.B.B. Sprout-Bauer, (717) 546-1517.

Conditions: 2,000 T/D of hardwood chips.

Cost (1990): $370,900 per machine. Four machines needed. Sprout-Bauer Model 45-1B.